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AN ELECTROCHEMICAL METHOD FOR REMOVAL OF PHOSPHATES FROM WASTE WATERS

Shafik E. Sadek

Dynatech Corporation
Cambridge, Massachuetts

February 1970

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"AN ELECTROCHEMICAL METHOD FOR REMOVAL OF PHOSPHATES FROM WASTE WATERS"

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by

Shafik E. Sadek

Dynatech Corporation
Cambridge, Massachusetts 02139

for the

FEDERAL WATER QUALITY ADMINISTRATION

DEPARTMENT OF THE INTERIOR

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ABSTRACT

Phosphates in waste water may be removed electrochemically utilizing sacrificial electrodes. The electrode metal is first dissolved by the flow of current then precipitates out, removing from solution the phosphate ions. This removal is either dependent on chemical reaction of the metal cation and the phosphate anions or, possibly, on the adsorption of the phosphate by the metal hydroxide floc.

Data on the phosphate removal was gathered using both aluminum and iron electrodes. Essentially complete removal was found to occur on using 300 coulombs/liter of charge flow with normal phosphate concentrations for both types of electrodes. Aluminum consumption averaged about 0.7 mass units per single mass unit of PO_4 removed for essentially complete phosphate removal. This mass ratio was about 2 for iron electrodes.

Treatment costs (excluding labor and filtration) have been estimated to be about 2.5 cents/1000 gal. and 8 cents/1000 gal. when using iron and aluminum electrodes respectively.

Exploratory tests indicated that flotation by means of the hydrogen generated during the electrolysis may be used to remove suspended solids from raw sewage while phosphates are being removed.

This report was submitted in fulfillment of Program No. 17010, Contract No. 14-12-405, between the Federal Water Quality Administration and Dynatech Corporation.

INTRODUCTION

In recent years, increased use of phosphate-based fertilizers and detergents have contributed to the large concentration of phosphates in waste water. This phosphate has been identified as one of the major causes of algae growth in receiving bodies of water, to the detriment of animal life (1,2). There is increasing need to treat the effluent from municipal treatment plants in order to reduce the phosphate content.

Use of electrochemical methods for treatment of waste waters is not new; several studies utilizing such methods have been described in patents and in the open literature (3,4,5,6). An experimental study has recently been reported in which several electrode materials were evaluated under a variety of experimental conditions to determine their effectiveness in removing various contaminants from waste water (7). It was demonstrated that the phosphate content of the effluent from a secondary treatment process could be reduced significantly. These experiments were not geared towards phosphate removal and were not run in a manner to remove phosphates most economically.

A study has been undertaken to evaluate an electrochemical method for removing phosphates. The objective of the program was to determine the effectiveness of an expendable electrode, direct current method for the removal of phosphates from the effluent of a secondary treatment process. The evaluation consisted of measuring electrode and power consumption and phosphate removal as a function of voltage, electrode material (aluminum or iron), electrode spacing, and residence time. Batch tests were followed by continuous flow (steady-state) tests. A number of tests were run on raw sewage to evaluate the use of the system as a combined phosphate removal/bubble flotation operation.

The economics of a conceptual plant were briefly evaluated and the cost of treating effluent under optimized conditions determined.

The term sacrificial electrode (e.g., aluminum or iron) implies that the metal composing the electrode dissolves forming positive ions. These ions react with the constituent anions of the solution (hydroxide, phosphate, etc.) and precipitate out as a floc. The result is removal of phosphates. At the cathode, the usual hydrogen evolution occurs and keeps the solution electrically neutral.

Tests with a non-sacrificial graphite anode showed that no phosphate was removed at similar current flows even though longer treatment periods were used (Table 1). The electrode material and the extent of phosphate removal are therefore intimately related to each other.

Phosphate removal by sacrificial electrode electrolytic techniques may be explained by either (or possibly both) of the following mechanisms:

1. Physical adsorption of the phosphates onto the hydroxide floc generated; or

2. Chemical reaction of the phosphate in the solution with the metallic ions followed by precipitation.

Choice of one or the other of the above-mentioned mechanisms to explain the phosphate removal phenomenon cannot presently be made with any degree of certainty owing to the lack of reliable data in the open literature on the solubility products of metallic phosphates. This report therefore does not presume to explain the mechanism of phosphate removal. It offers the results of a systematic set of tests which allows the evaluation of the sacrificial electrode process for phosphate removal on a sounder basis than was possible earlier.

Table 1

EFFECT OF ELECTRODE MATERIAL ON PHOSPHATE REMOVAL

Electrode Material	Potential (Volts)	Current (Amps)	Time (Hours)	Initial PO_4 (mgm/liter)	Final PO_4 (mgm/liter)
Carbon	5	0.04	1.5	30.5	27.8
Carbon	15	0.13	1.3	30.5	29.3
Aluminum	5	0.14	0.5	30.5	0.2

EXPERIMENTAL

TEST PROCEDURE

Three sets of tests were performed. The first set consisted of batch tests performed in a Plexiglas cell 4 in. wide by 6-1/2 in. long and 8 in. high (Figure 1). The electrode spacing was made variable by supporting the electrodes externally. The dead space behind the electrodes was reduced by filling it with Plexiglas blocks. A small laboratory stirrer was used to keep the contents of the cell mixed.

The second set of tests consisted of flow tests in single electrode-pair systems. Two such cells (Figure 2) were assembled with nominal electrode spacings of 1/2 in. and 1 in. In these two sets, smooth aluminum and iron electrodes were used; each electrode measured 4 in. x 4 in. No external stirring was used.

The third and final set of tests were run in a multiple electrode (aluminum) system (Figure 3). Electrode spacings were 1 in. and each electrode was 6 in. x 6 in. of aluminum screen. Here, separation of the floc from the solution was accomplished by bubble flotation. No external stirring was used.

The solutions used in these tests were either "synthetic effluent" or true secondary treatment effluent. The synthetic effluent was used as a convenience to determine the trends in the performance as a function of the system parameters. This was a natural choice as it was simple to reproduce and did not deteriorate with time. True effluent was used to check the trends determined in the synthetic effluent tests. The synthetic effluent contained 0.25 gm/liter sodium chloride, 0.41 gm/liter sodium bicarbonate and dibasic sodium phosphate to give 20, 40, 80, and 300 mgm/liter PO_4 as required.

Raw sewage was used in the multiple electrode system in order to determine the effectiveness of bubble flotation in reducing suspended impurity concentrations in conjunction with phosphate removal.

The phosphate measurement was made according to the procedures recommended in "Standard Methods for the Examination of Water and Waste Water (8).

Figure 1
PHOTOGRAPH OF BATCH CELL

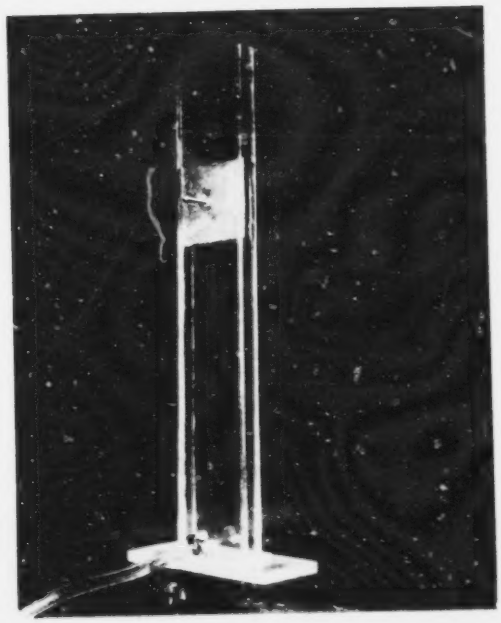

Figure 2
PHOTOGRAPH OF FLOW CELL

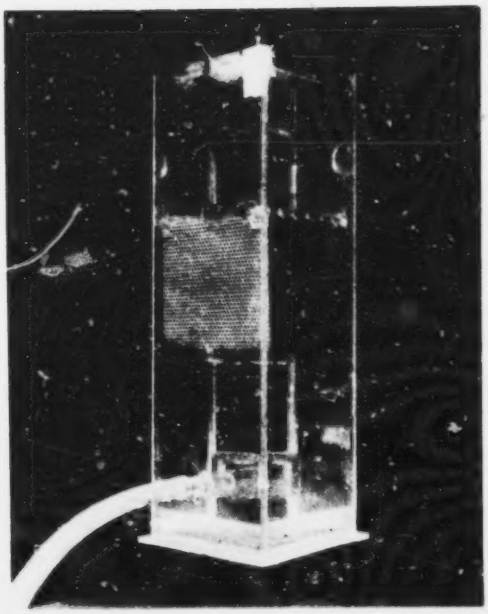

Figure 3

PHOTOGRAPH OF LARGE FLOW CELL

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In both synthetic and secondary effluents, all results are based on an orthophosphate measurement. Preliminary analysis showed that total and orthophosphate content of secondary effluent were essentially the same.

The test procedure consisted of starting with a known phosphate content in the solution, adjusting the cell potential to the desired level, and checking the current as the run proceeded. At the end of the run, the phosphate content was measured.

In some of the batch tests, the reduction in anode weight was also measured in order to determine the current efficiency with respect to the anode reaction. In continuous flow tests, the inlet and outlet phosphate concentrations and current were measured during steady state conditions.

TEST RESULTS

Test results consist of the current as a function of the system parameters, electrode consumption rates, and phosphate removals.

Current

The current (I) depends on the following system parameters:

- voltage (V)
- solution conductivity (σ)
- electrode spacing (s)
- electrode area (A)
- electrode material

Here, it is assumed that edge effects are negligible. For the liquid between the electrodes, one may relate the above parameters as

$$\left(\frac{I}{V}\right)_{\circ} = \frac{\sigma A}{s}$$

where the subscript "$\circ$" denotes absence of electrode and polarization potentials. Since the latter potentials are not negligible, the observed current will be less than

that calculated from the above equation using measured voltage. The ratio of observed to calculated current,

$$\eta = \frac{I}{I_o} ,$$

is therefore less than unity and represents the current effectiveness at a given voltage.

Figure 4 shows I/σ, current divided by conductivity, of different effluents plotted against the operating voltage in the batch cell at an electrode spacing of 1.6 cm. The current remained essentially constant throughout the runs and no change in conductivity was noticed. Aluminum electrodes were used. One set consisted of tests with aluminum foil electrodes; the other with perforated aluminum sheets. It is seen that at a given voltage the current is higher for the perforated sheets than it is for the smooth electrode. This is probably due to the higher surface area of the screen type electrode.

The current effectiveness, η, is shown also plotted against the voltage (Figure 5). This effectiveness is only slightly dependent on the voltage and equal to about 0.13 in the case of the foil electrode and about 0.2 in the case of the perforated sheet electrode within the range of interest.

The effect of electrode spacing on the efficiency η is shown in Figure 6. These tests were run with synthetic sewage in the batch cell, using perforated sheet electrodes. The results indicate that smaller spacings give a slightly higher value of η than wider ones. This is probably due to the higher current densities through the system at a given voltage when the spacing is reduced. This increase in current results in an increase in the stirring of the system, owing to the greater bubble generation rate. The value of η also increases with an increase in operating voltages. This too is probably caused by the higher current accompanied by an increase in stirring of the system.

Replotting the data as η against the current (I) instead of the voltage (V) correlates the data with considerably less scatter (Figure 7). A better correlation is however to be expected as both ordinate and abscissa are proportional to the current.

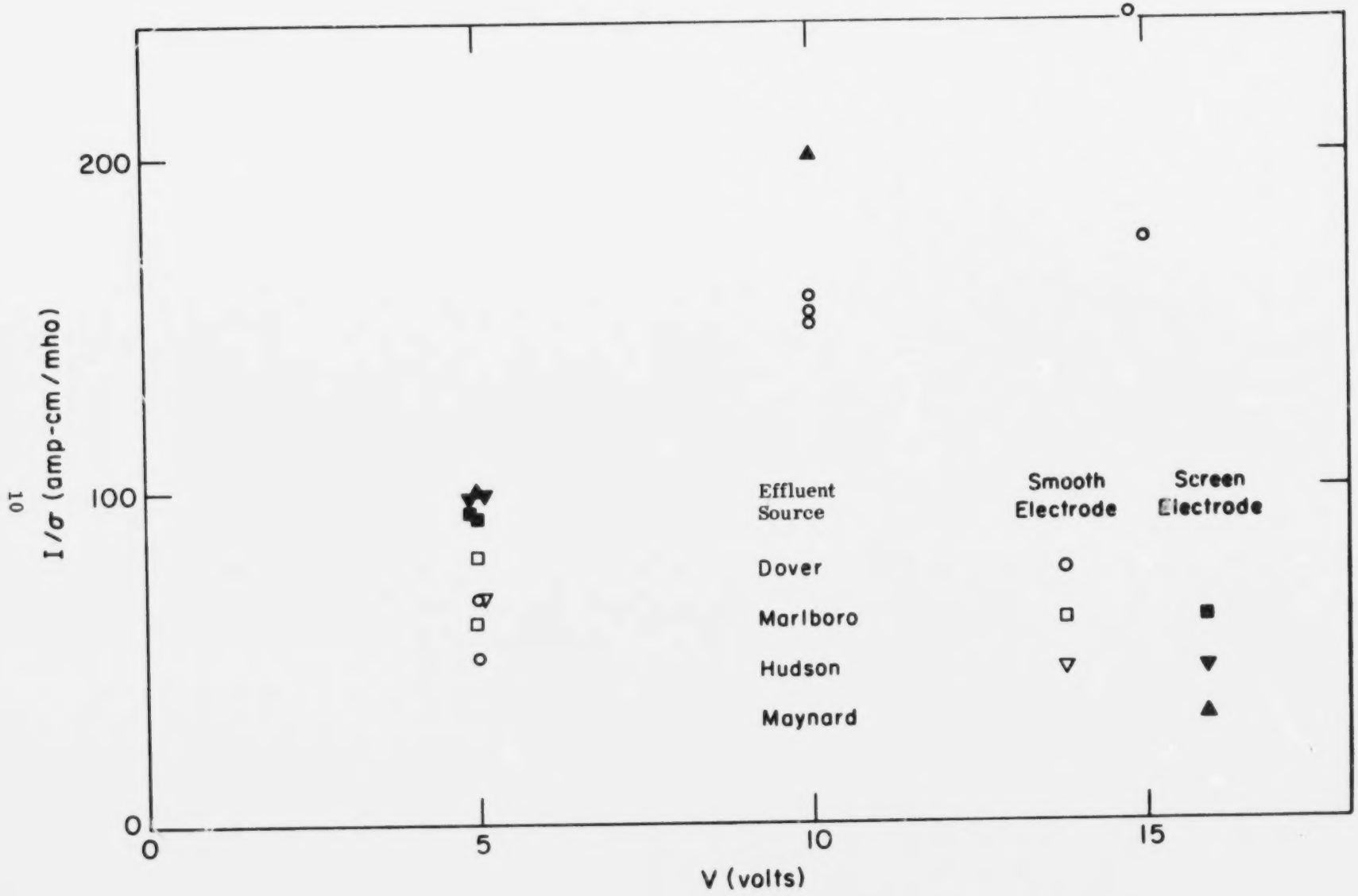

Fig. 4 Current-Voltage Relationship Using Aluminum Electrodes (Electrode Spacing = 1.6 cm)

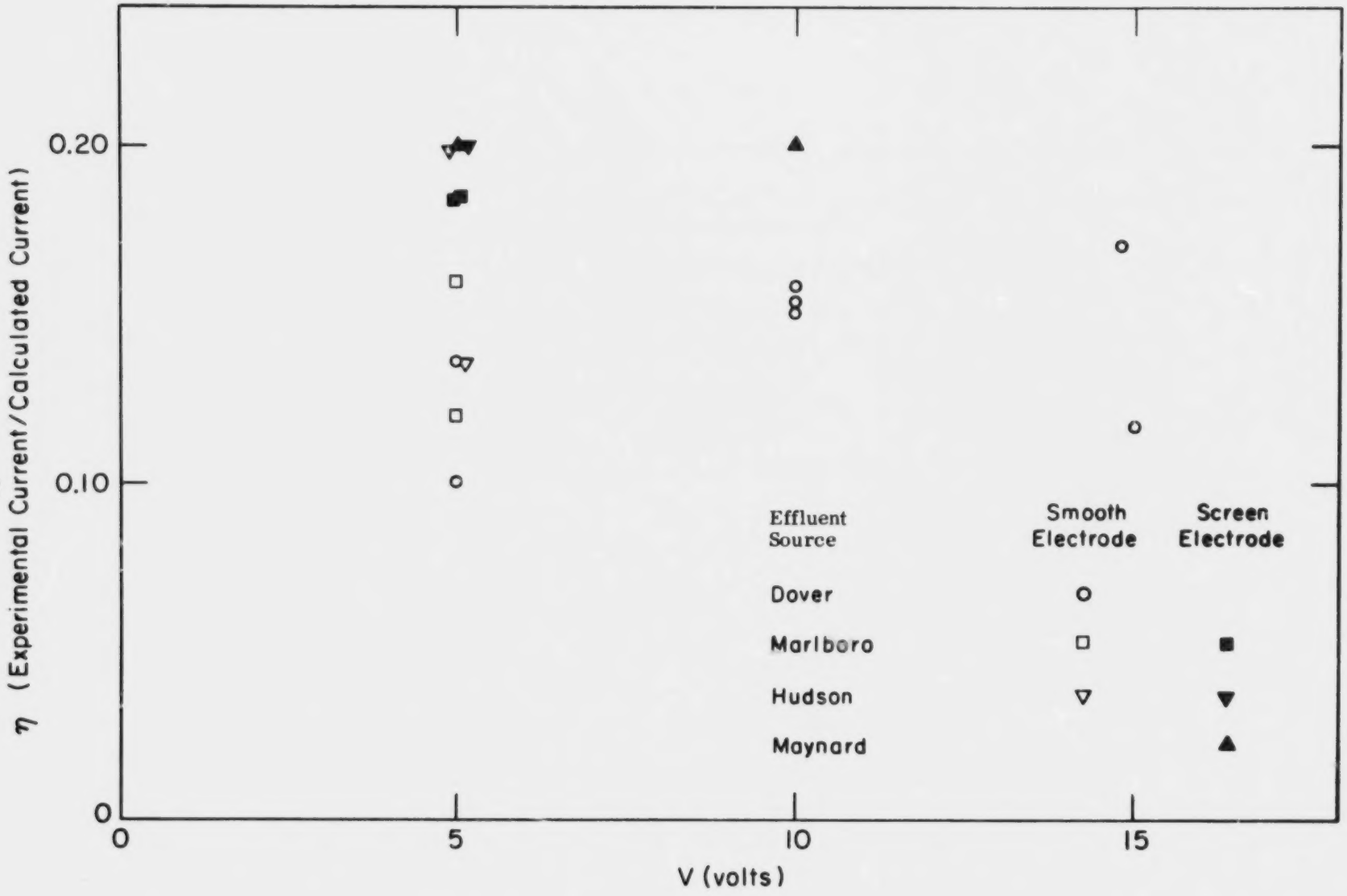

Fig. 5. Current Flow Effectiveness Using Aluminum Electrodes (Electrode Spacing = 1.6 cm)

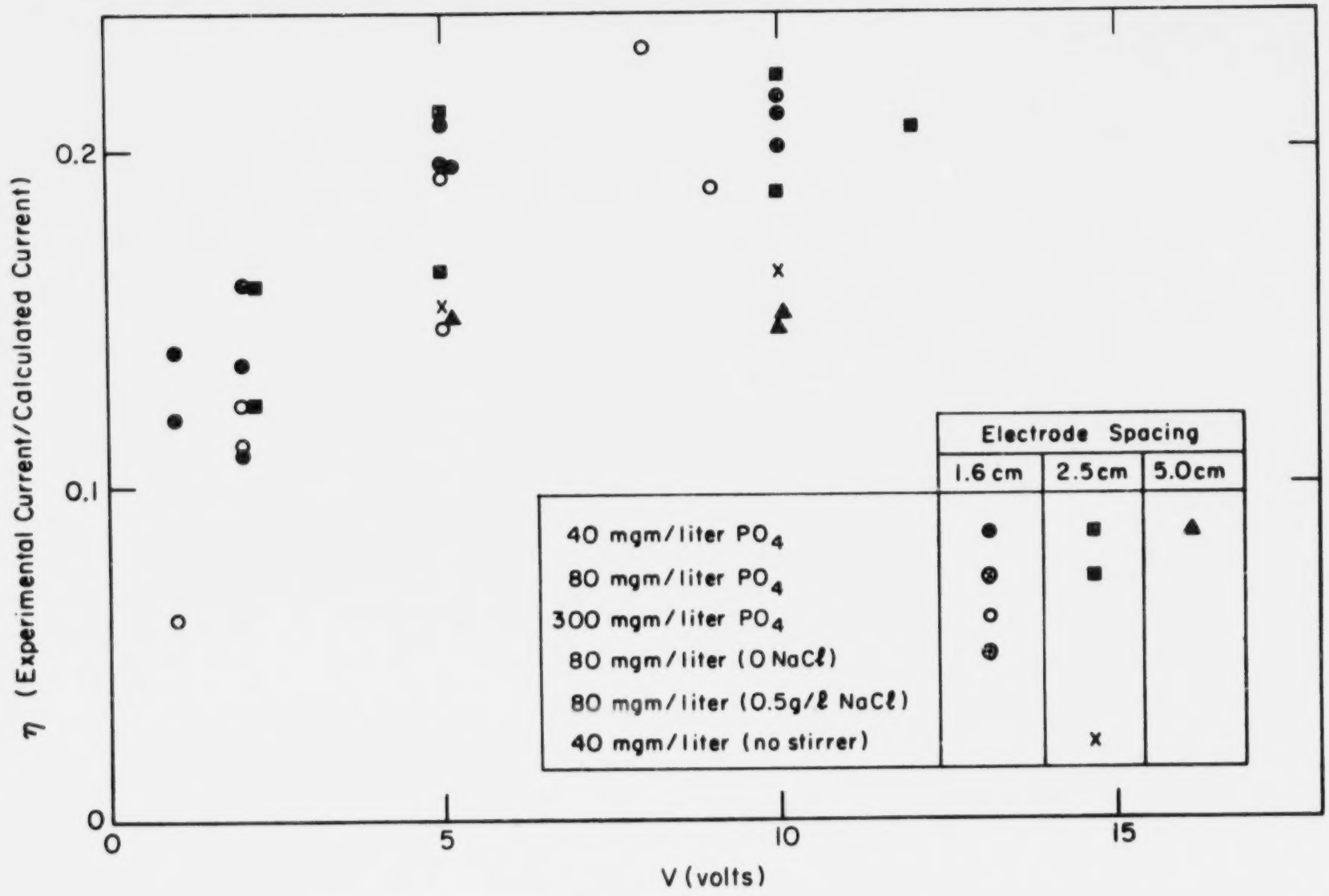

Fig. 6. Effect of Electrode Spacing on Current Effectiveness Using Aluminum Electrodes

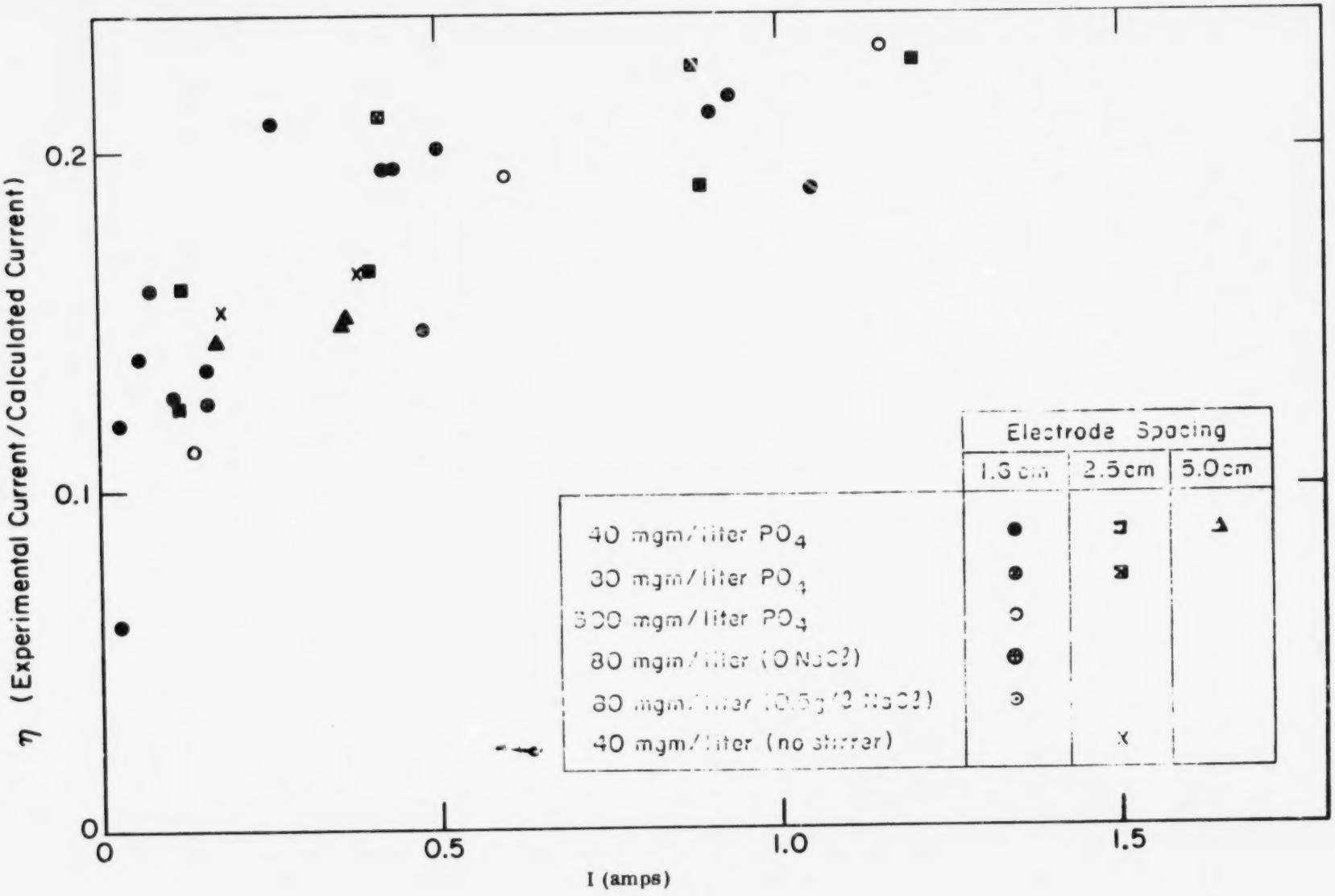

Fig.7 Current Flow Effectiveness as a Function of Current (Aluminum Electrodes)

No tests were run to determine the effect of electrode height on the current. Even though it is probably that such an effect does exist (since height will affect degree of turbulence), it may be expected that this effect is small.

The effect of stirring on the current was found to be small in the case of runs with aluminum electrodes (Figure 7).

With iron electrodes, the same trends were noted (Figures 8 and 9). In this case, synthetic effluent showed slightly lower current effectiveness than secondary treatment effluents from Marlboro or Hudson, Massachusetts.

Electrode Consumption

Electrode dissolution rates were determined experimentally. This was accomplished by using thin foil in the case of aluminum and 0.006 in. thick iron sheet as electrodes, and weighing them before and after the tests. On comparing these weight losses to the current, it was shown that under all of the operating conditions used, the current efficiency scattered around 100% with respect to aluminum dissolution (i.e., one Faraday dissolved one equivalent). Side reactions (oxygen evolution) may therefore be considered negligible.

When iron electrodes are used, the current efficiency appeared to be 100% with respect to the Ferrous iron. Since no gas was observed to evolve it, it is probable that the electrode did, in fact, dissolve as Fe^{++}.

Data supporting these current efficiencies are listed in Tables 2 and 3.

Phosphate Removal

Phosphate removal is intimately related to the quantity of sacrificial electrode material introduced into the solution. Three steps are necessary for the phosphate removal to occur:

1. Metal ions (or floc) must first be generated;

2. The phosphate and the metal ion (or the floc) must be transferred from their respective high concentration regions to the low concentration "reaction" zone;

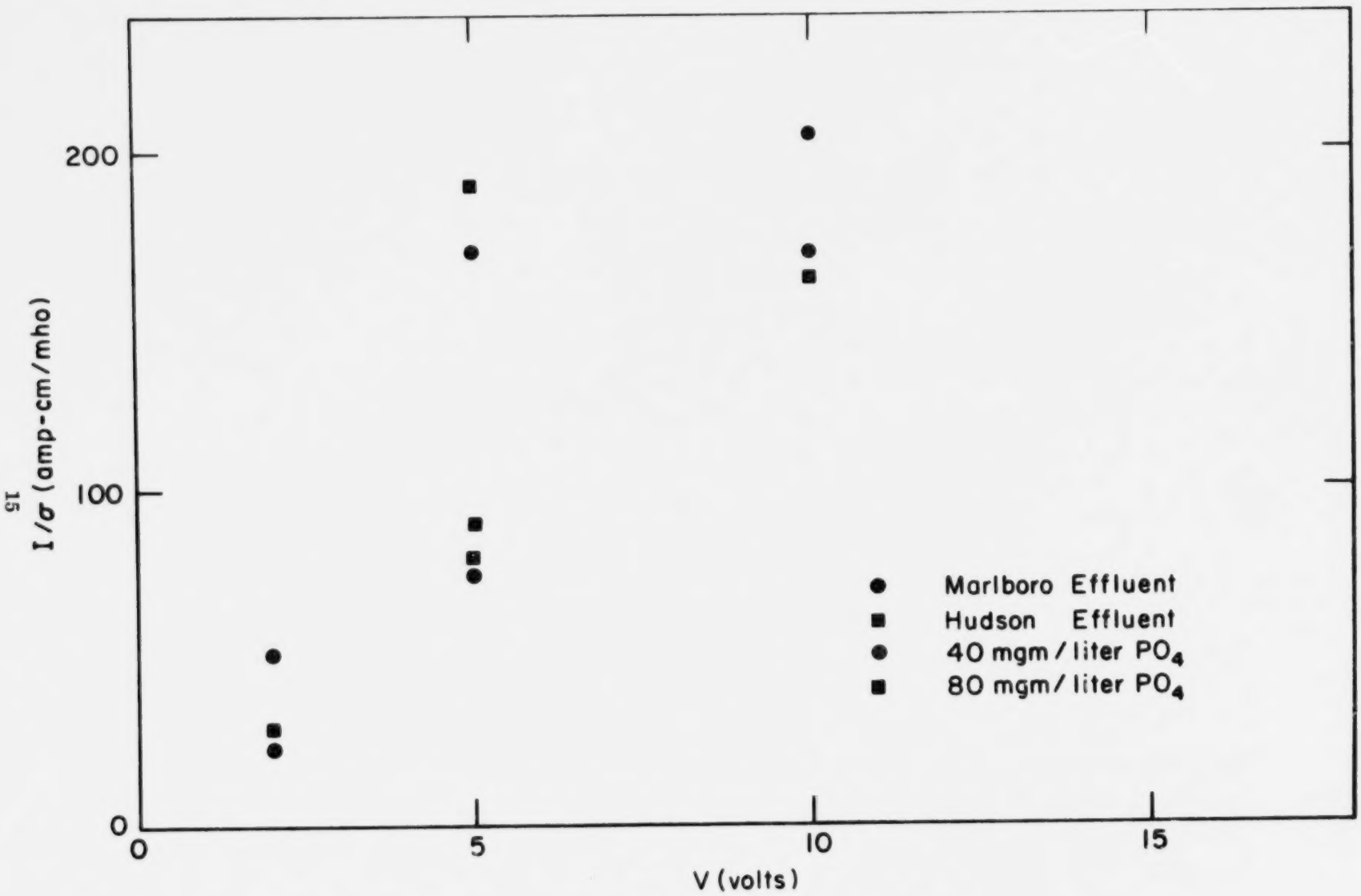

Fig. 8 Current-Voltage Relationship Using Iron Electrodes (Electrode Spacing = 1.6 cm)

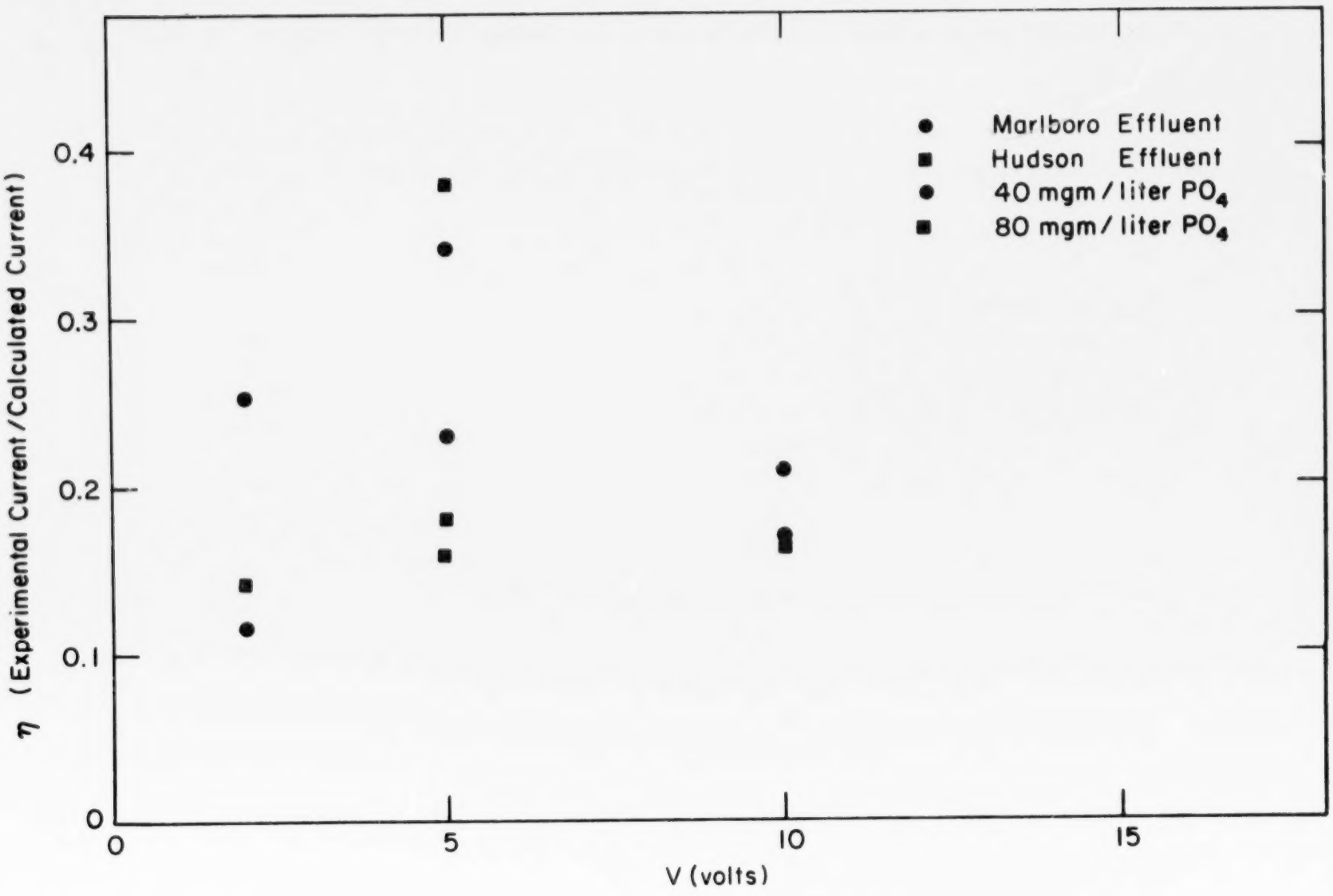

Fig. 9. Current Effectiveness Using Iron Electrodes

Table 2

CURRENT EFFICIENCY WITH RESPECT TO
ALUMINUM ELECTRODE DISSOLUTION

Run	Sewage Origin	Potential (volts)	Current (amps)	Time (min.)	Calculated Anode Wt. Loss (mgm) (W_c)	Measured Anode Wt. Loss (mgm) (W_m)	Electrode Dissolution Current Efficiency (ϵ) = W_m/W_c)
4R	Dover, N. H.	10	0.3	5	8.4	18.8	2.19
5R	Dover, N. H.	10	0.35	10	19.6	23.9	1.22
5RR	Dover, N. H.	10	0.35	15	29.4	31.8	1.08
5W	Dover, N. H.	10	0.3	10	16.8	18.9	1.13
6R	Portsmouth, N.H.	5	0.95	15	80	109.9	1.37
7	Portsmouth, N.H.	5	0.90	10	50.3	65.9	1.31
8	Portsmouth, N.H.	5	0.90	5	25.2	32.9	1.31
2-3	Marlboro, Mass.	5	0.15	5	4.2	2.5	0.60
2-7	Hudson, Mass.	5	0.15	5	4.2	3.8	0.90
15	Marlboro, Mass.	5	0.17	10	9.5	9.0	0.95
16	Hudson, Mass.	5	0.16	10	9.0	10.9	1.21
19	Dover, N. H.	10	0.315	10	17.2	19.8	1.15
20	Dover, N. H.	5	0.135	10	7.6	7.6	1.0
21	Dover, N. H.	5	0.130	1	7.3	9.0	1.23
22	Dover, N. H.	10	0.29	1	1.6	1.3	0.81
23	Dover, N. H.	14.8	0.48	1	2.7	3.5	1.30
24	Dover, N. H.	5	1.0	1	5.6	5.0	0.89
17	Dover, N. H.	10	0.35	10	19.6	19.3	0.98
18	Dover, N. H.	10	0.32	10	17.9	19.3	1.08
130	Synthetic	5	0.30	5	8.4	7.7	0.92

Table 3

CURRENT EFFICIENCY WITH RESPECT TO IRON ELECTRODE DISSOLUTION

Run	Sewage Origin	Potential (volts)	Current (amps)	Time (min.)	Calculated No. of Equivalents Dissolved x 1000 (n)	Measured Electrode wt. Dissolved (mgm) (m_c)	Valency of Metal Dissolved $= \frac{55.85xn}{m_c}$
137	Marlboro	2	0.1	5	0.311	13.0	1.34
92	Synthetic	10	0.94	15	8.77	249.2	1.97
94		2	0.12	5	0.373	12.2	1.71
95		5	0.38	5	1.18	30.0	2.20
96		10	0.9	5	2.80	82.9	1.89
107		5	0.39	10	2.42	71.9	1.88
108		5	0.38	25	5.91	69.3	4.76
113		2	0.11	120	8.21	216.7	2.12
120		2	0.13	60	4.85	129.6	2.09

3. When phosphate and metal ion (or floc) come into contact, they must then react (chemically or by adsorption). At a given overall concentration of both constituents this reaction proceeds to an equilibrium point beyond which no more reaction will occur.

It was suspected at the beginning of the program that phosphate removal may occur during the electrolytic process by some nucleation mechanism involving the precipitation of calcium and magnesium phosphates. This was disproved, however, when it was shown that the calcium and magnesium contents of the effluent remained unchanged during the process (Table 4). (Note too that ammonia nitrogen also remains unchanged but that TOC content is reduced slightly.)

Some tests were performed to determine the extent to which each of the above steps govern the overall phosphate removal rate. It was shown (see Figure 6) that variations in the degree of stirring affected the phosphate removal only to a small degree, at least in the case of aluminum electrodes. Mass transfer, i.e., step 2, is therefore not limiting the rate of removal appreciably: it is probable that the rising hydrogen bubbles at the cathode generated sufficient circulation in the process so that the presence of a mechanical stirrer had little additional effect on the removal rate.

Other tests showed that after the voltage was switched off, no further reduction in phosphate content could be detected--even on stirring the mixture for periods of up to one hour (Table 5). Step 3, the reaction (chemical or adsorption) rate, is therefore not limiting the phosphate removal rate.

The removal rate must therefore be limited by the rate of ion or floc generation, while the extent to which the removal proceeds must be governed by some equilibrium relationship within the mixture.

This would indicate that the extent of phosphate removal from a given effluent will depend only on the mass of metallic ion or floc generated and the phosphate concentration. Since the mass of metal dissolved is directly proportional to the charge flow through the system, one may expect that the extent of phosphate removal

Table 4

EFFECT OF ELECTROLYSIS ON POLLUTANTS IN MAYNARD, MASSACHUSETTS EFFLUENT

Sample	PO_4 mgm/liter	Ca mgm/liter	Mg mgm/liter	Alkalinity ($CaCO_3$) mgm/liter	Total Organic Carbon TOC mgm/liter	NH_3-N mgm/liter
Untreated	33	11	2.8	170	25.0	36.0
5 volts, 5 mins.	14	11	2.8	162	18.0	34.6
10 volts, 5 mins.	5	8	2.5	167	16.4	36.8

Table 5

EFFECT OF TIME ON PHOSPHATE REMOVAL
AFTER VOLTAGE HAS BEEN SWITCHED OFF

Time* (mins.)	PO_4 (mgm/liter)
~0	3
15	2.9
30	2.9
60	3.0

*Time indicates the period starting from the instant the voltage was switched off.

Voltage = 5 volts, applied for 5 mins.

Initial PO_4 content ≅ 40 mgm/liter

from a given solution in a certain cell may be correlated with the charge flow through the system. Variations in voltage, or residence time will affect the removal only inasmuch as they affect the total charge flow through the system.

Figure 10 shows the phosphate content of synthetic sewage samples, initially containing about 40 and 75 mgm/liter PO_4 as a function of the charge flow through the cell when using aluminum electrodes. Runs with different voltages and residence times correlate well in this set with an electrode spacing of 1.6 cm.

Going one step further, one may expect to correlate the data gathered in different cell geometries by plotting the residual concentration of phosphate against the total mass of electrodes dissolved per unit volume of solution. In other words, it is possible to plot the residual phosphate concentration of a certain solution against the charge flow per unit volume of solution, irrespective of the voltage, cell spacing, or residence time used in the test; these last three variables affect the process only inasmuch as they affect the total charge flow.

The data represented in Figures 11, 12, and 13 show the concentration of phosphate as a function of the charge flow per unit volume through the cell; the curves gather the data for various electrode spacings when utilizing aluminum electrodes and for three different initial phosphate concentrations, 40, 80, and 300 mgm/liter PO_4. Data for secondary treatment effluents from the Maynard and Marlboro, Massachusetts, plants are also shown in Figure 11. Their initial ortho (and total) phosphate concentrations were 33 and 41 mgm/liter PO_4 respectively. The drop in concentration is appreciably greater at the higher concentrations, but the relative reduction is lower (i.e., fraction removed based on initial content). This is in keeping with the concept of an equilibrium between the absorbed and the dissolved phosphate.

A similar approach may be used to correlate the flow tests. Here, too, one may plot the effluent phosphate concentration against the charge per unit volume, which in this case is equal to the current divided by the flow rate through the system. Figures 14, 15, and 16 show the data generated in the single electrode-pair system. For comparison, the same figures show the curves describing the data generated in the batch cells (for ~40 and ~80 mgm/liter initial PO_4 content).

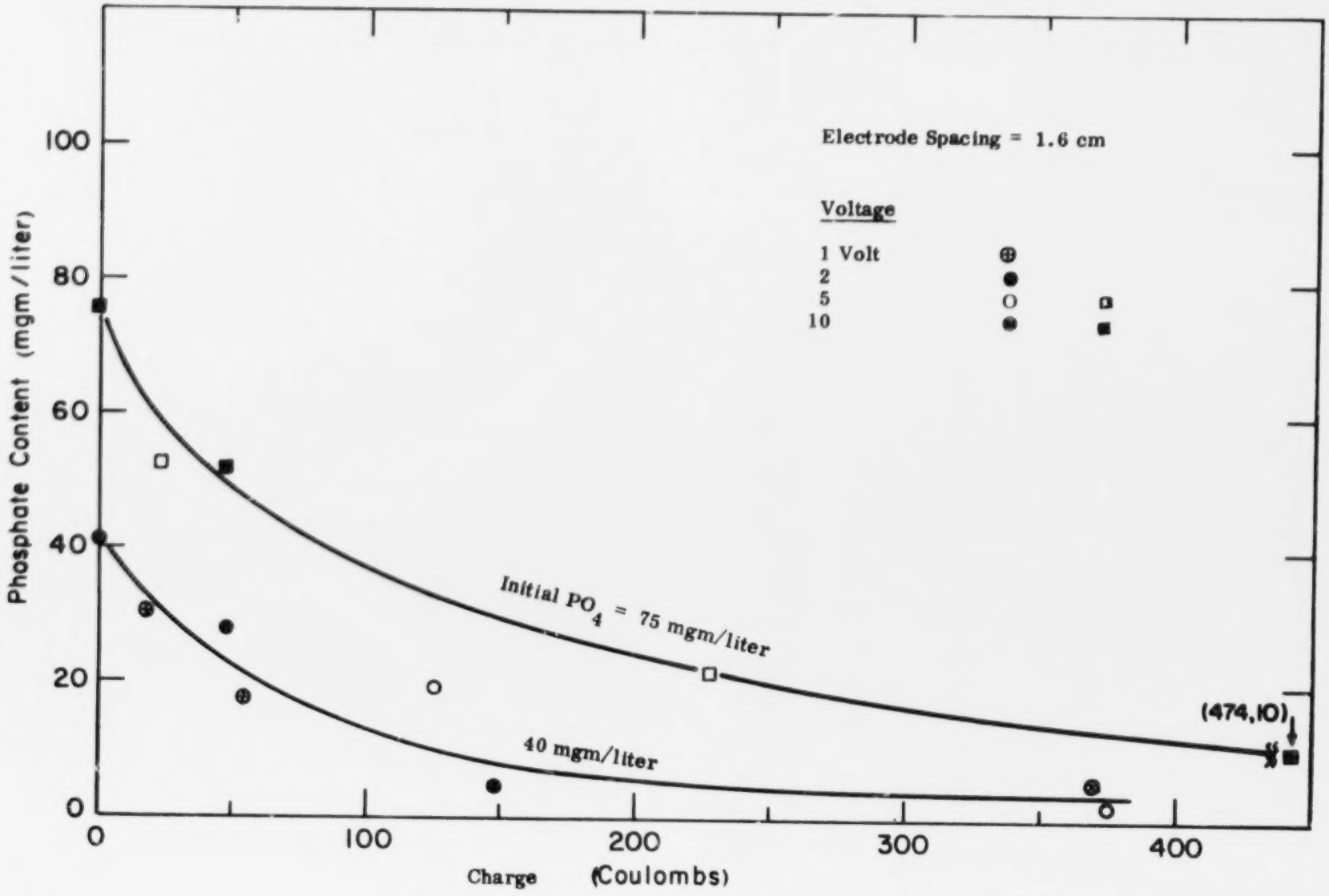

Fig. 10. Effect of Charge Flow on Phosphate Removal Using Aluminum Electrodes

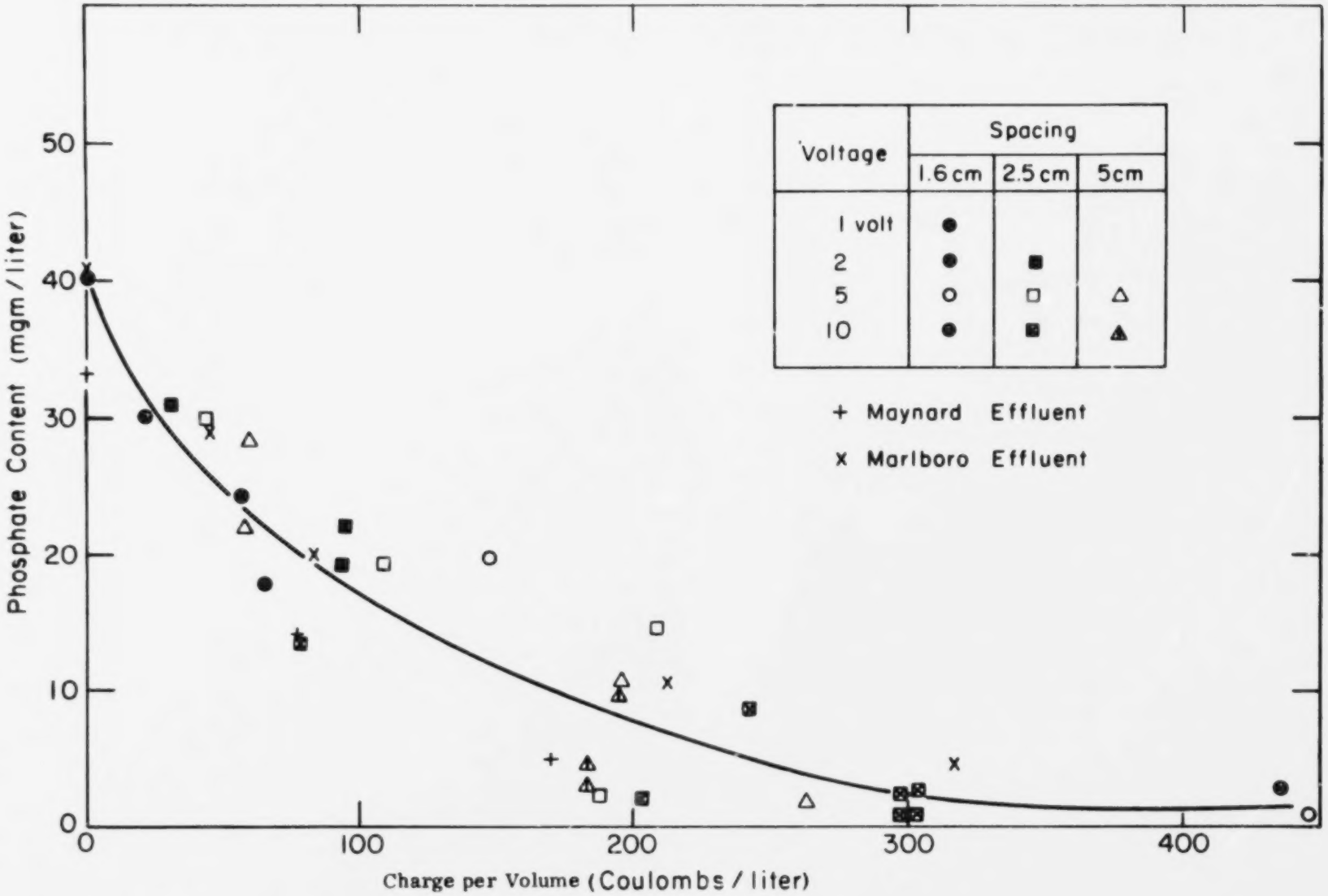

Fig. 11. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Batch Cell

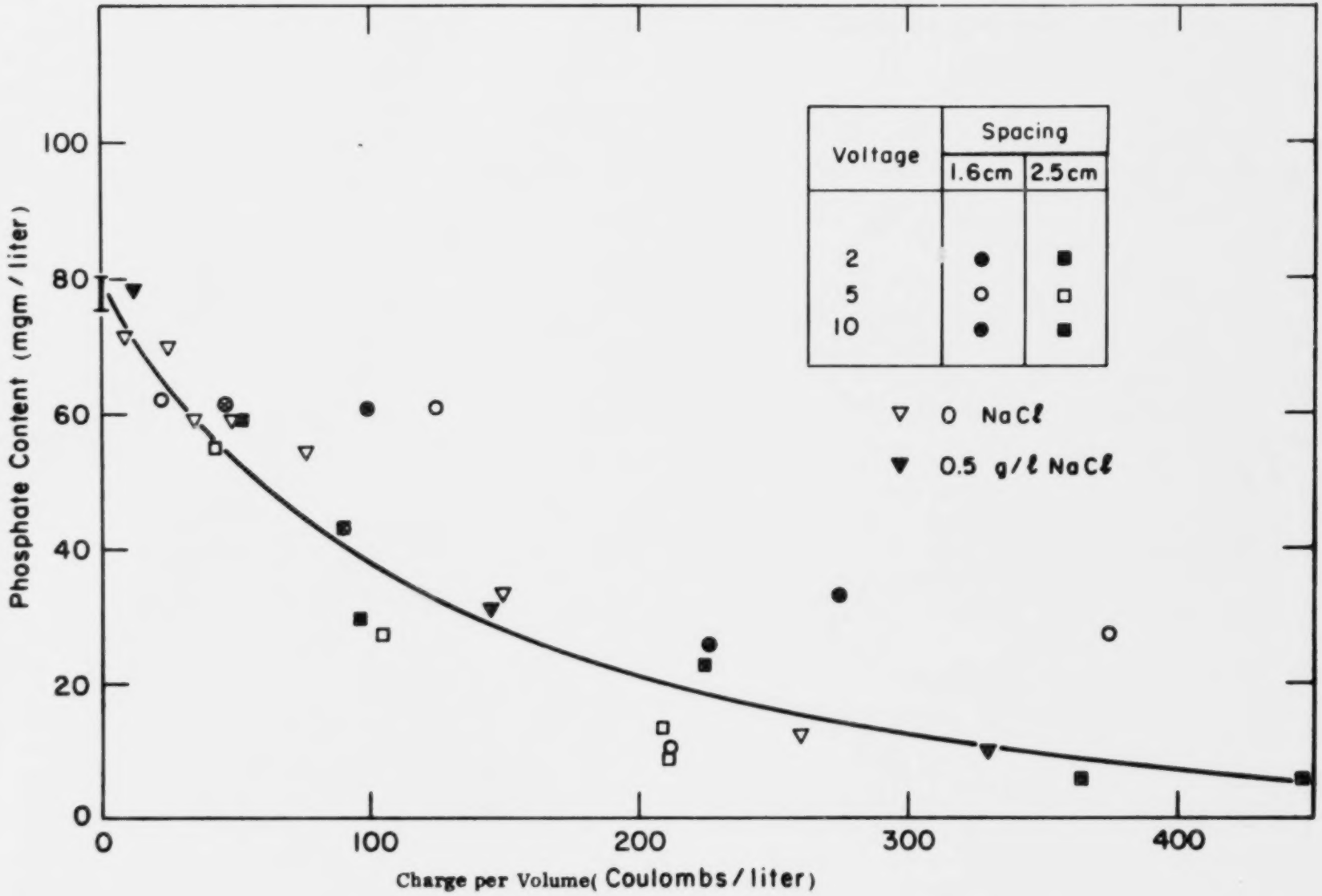

Fig. 12. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Batch Cell

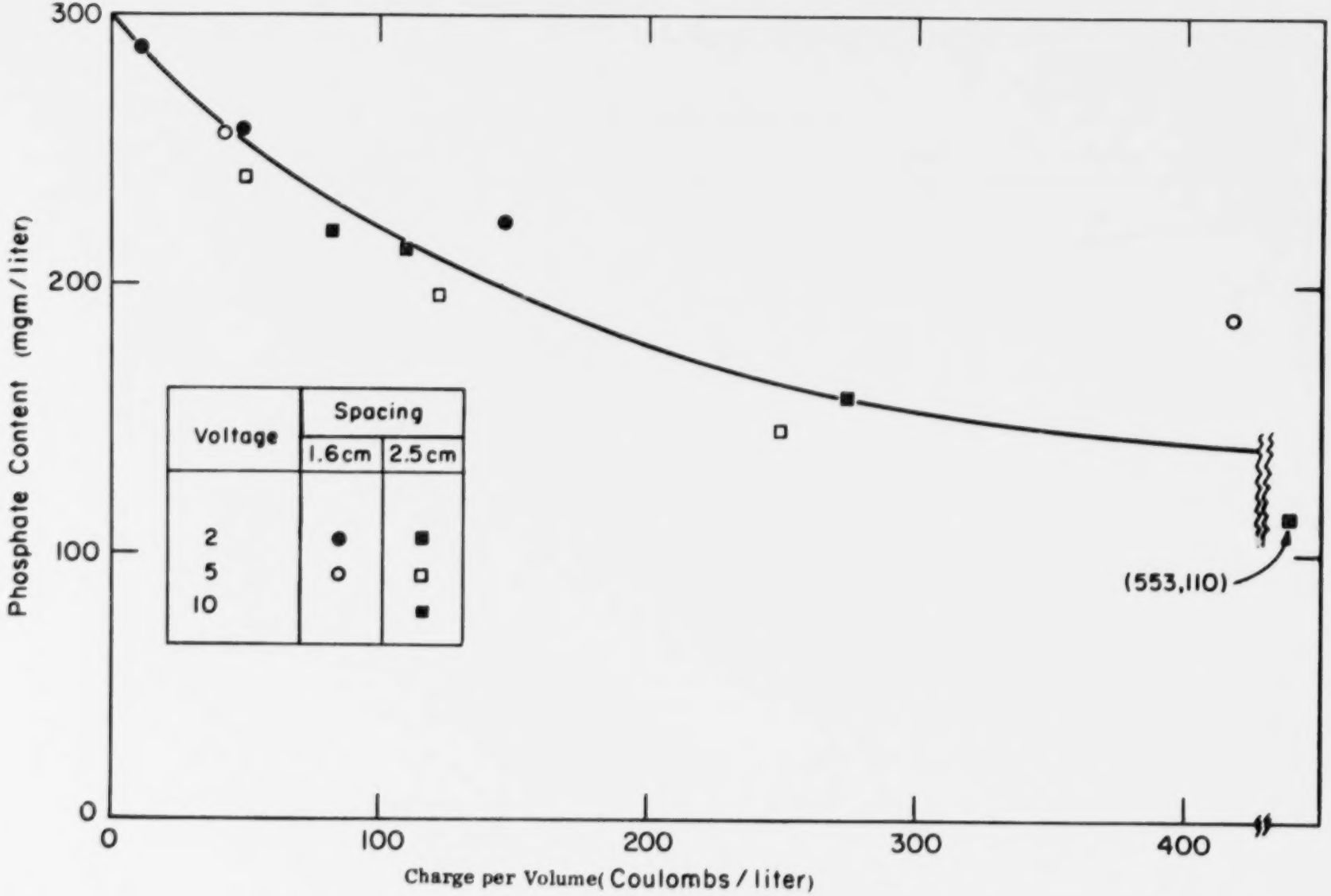

Fig.13. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Batch Cell

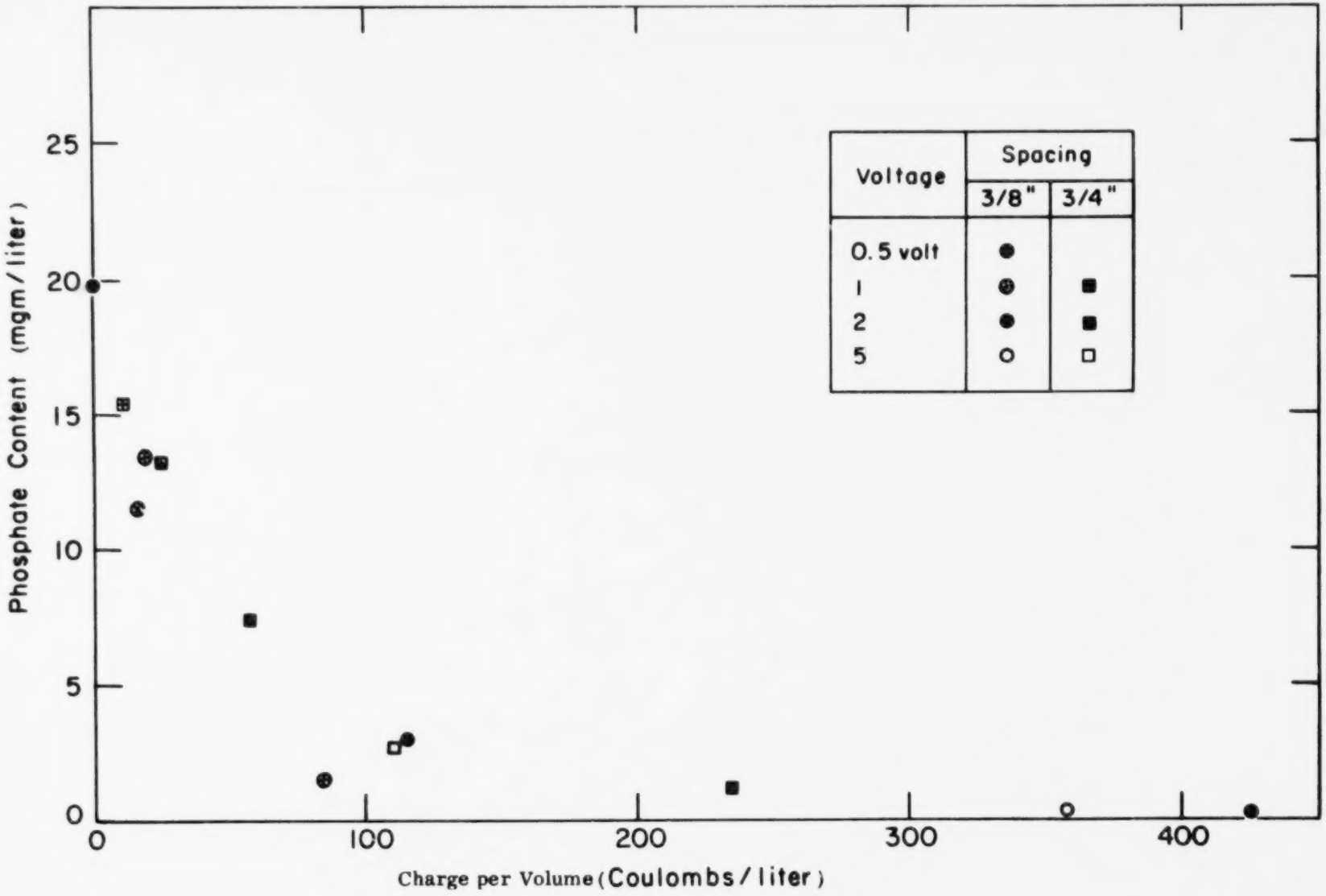

Fig. 14. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Flow Cells

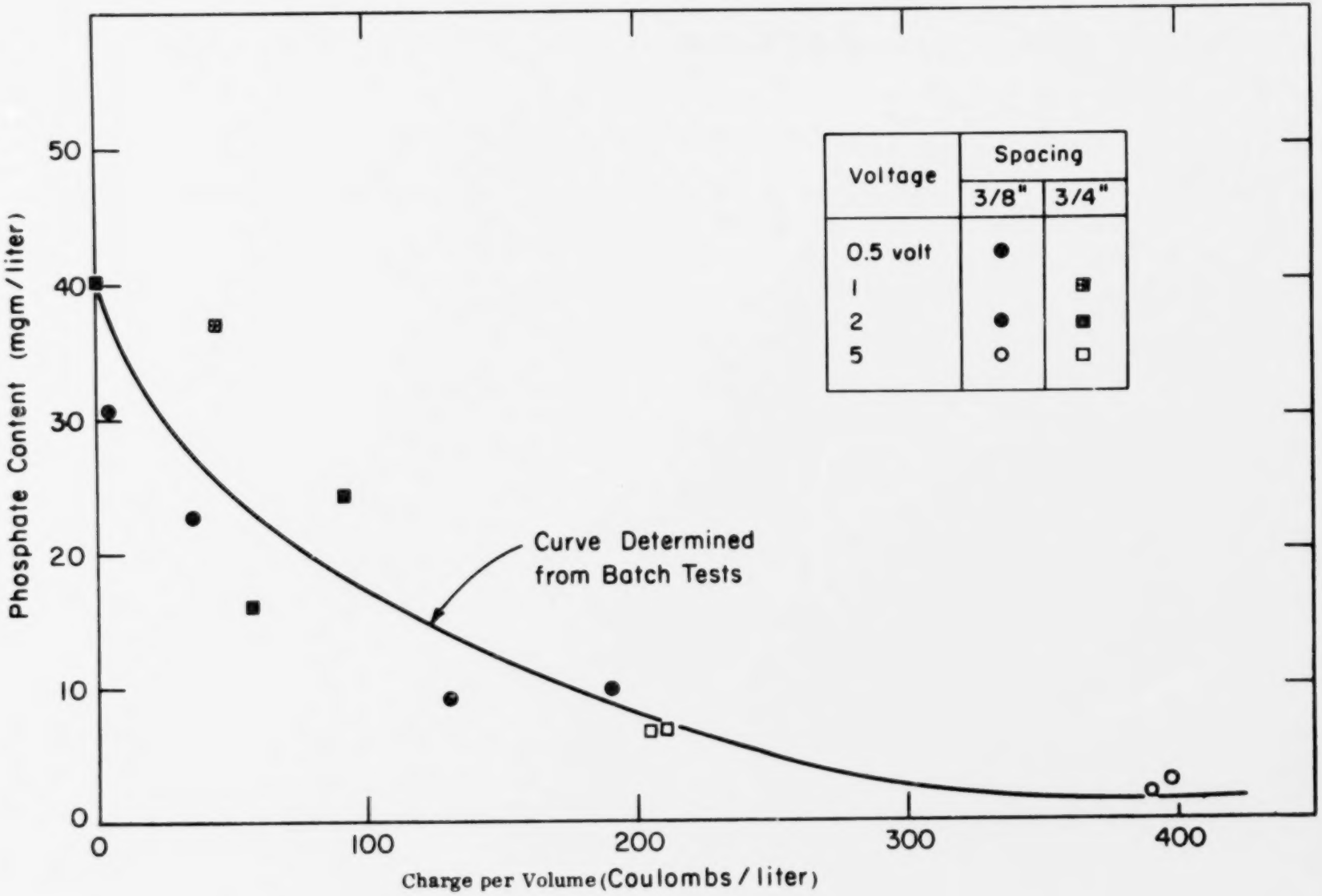

Fig. 15. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Flow Cells

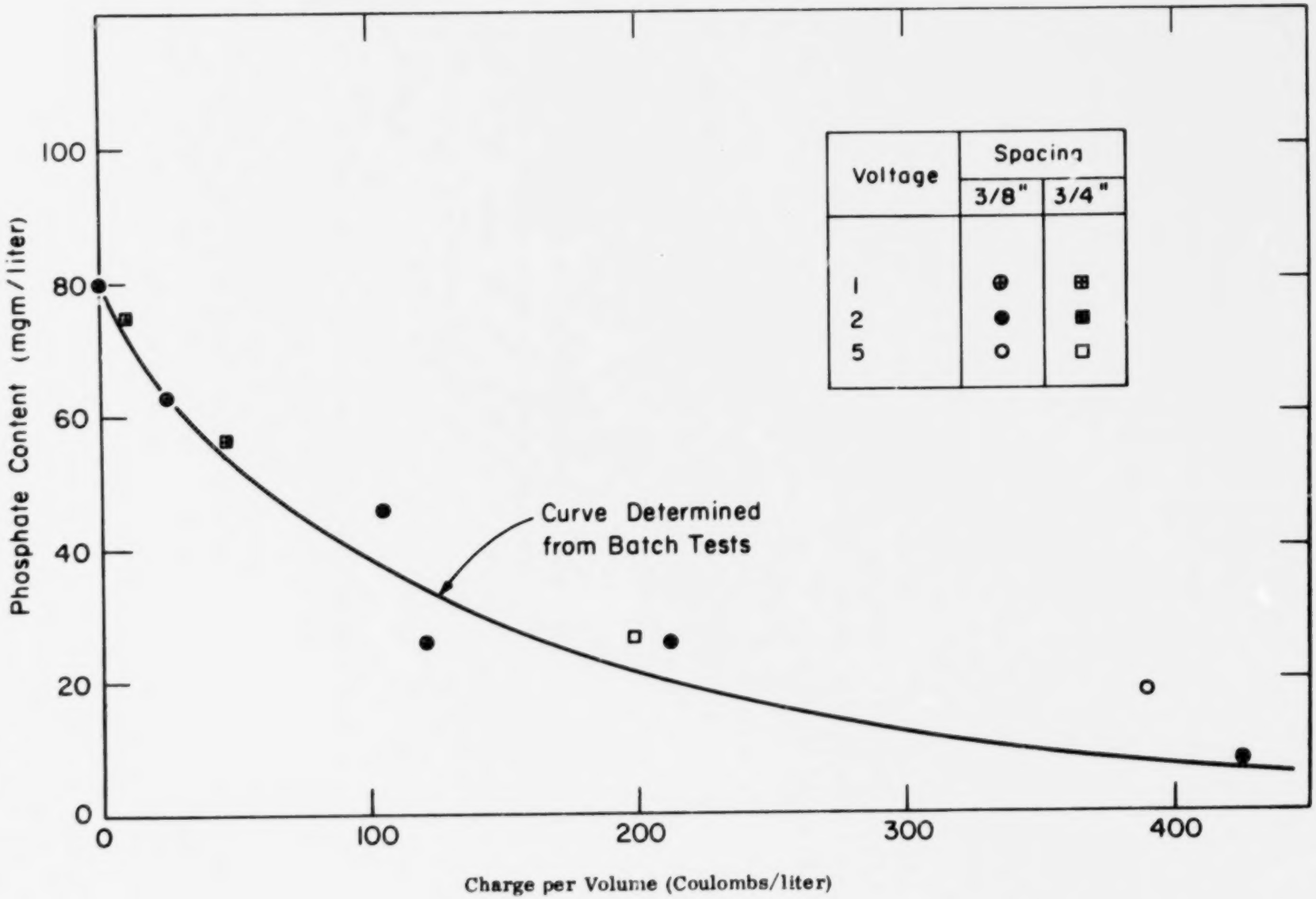

Fig. 16. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Flow Cells

The data generated in the multiple electrode system (Figure 17) falls slightly above the data generated in the other systems. The explanation for this discrepancy lies in the facts that some unavoidable channeling exists and that current flows are not exactly equally distributed between the electrodes owing to some non-uniformity in their assembly. The overall performance of the system is therefore poorer than may be expected.

The curves of phosphate content versus charge flow are shown cross-plotted as the mass ratio of aluminum dissolved to phosphate removed (assuming 100% current efficiency) against the concentration of phosphate removed for four initial phosphate concentrations of 20, 40, 80, and 300 mgm/liter PO_4 (see Figure 18). Ninety-five per cent PO_4 removal requires an aluminum consumption of about 0.6 - 0.8 mass units per unit of PO_4 removed, the actual value depending on the initial phosphate content.

Iron electrodes behaved in a manner similar to aluminum electrodes. Figure 19 shows the data gathered on synthetic and real effluents containing initially about 20 mgm/liter PO_4. Both batch and flow test results are plotted in this figure. Figures 20 and 21 show the data gathered on synthetic effluents initially containing 40 and 80 mgm/liter PO_4.

Comparing these results to those obtained with aluminum electrodes, there appears to be a slightly lower charge flow requirement with iron than with aluminum. Mass consumption of the iron electrodes (Figure 22) is, however, considerably greater than the aluminum electrodes owing to the greater equivalent weight of iron.

Raw Sewage Tests

Three sets of tests using raw sewage were run in the multiple electrode flow cell. The results of these tests are listed in Table 6.

Raw sewage is introduced at the top of the cell and flows downwards through the electrode array. Bubbles generated at the cathode (hydrogen) attach themselves to the suspended matter and float it upwards to the top of the cell where it can be skimmed off. The effluent is considerably clearer than the sewage introduced. A

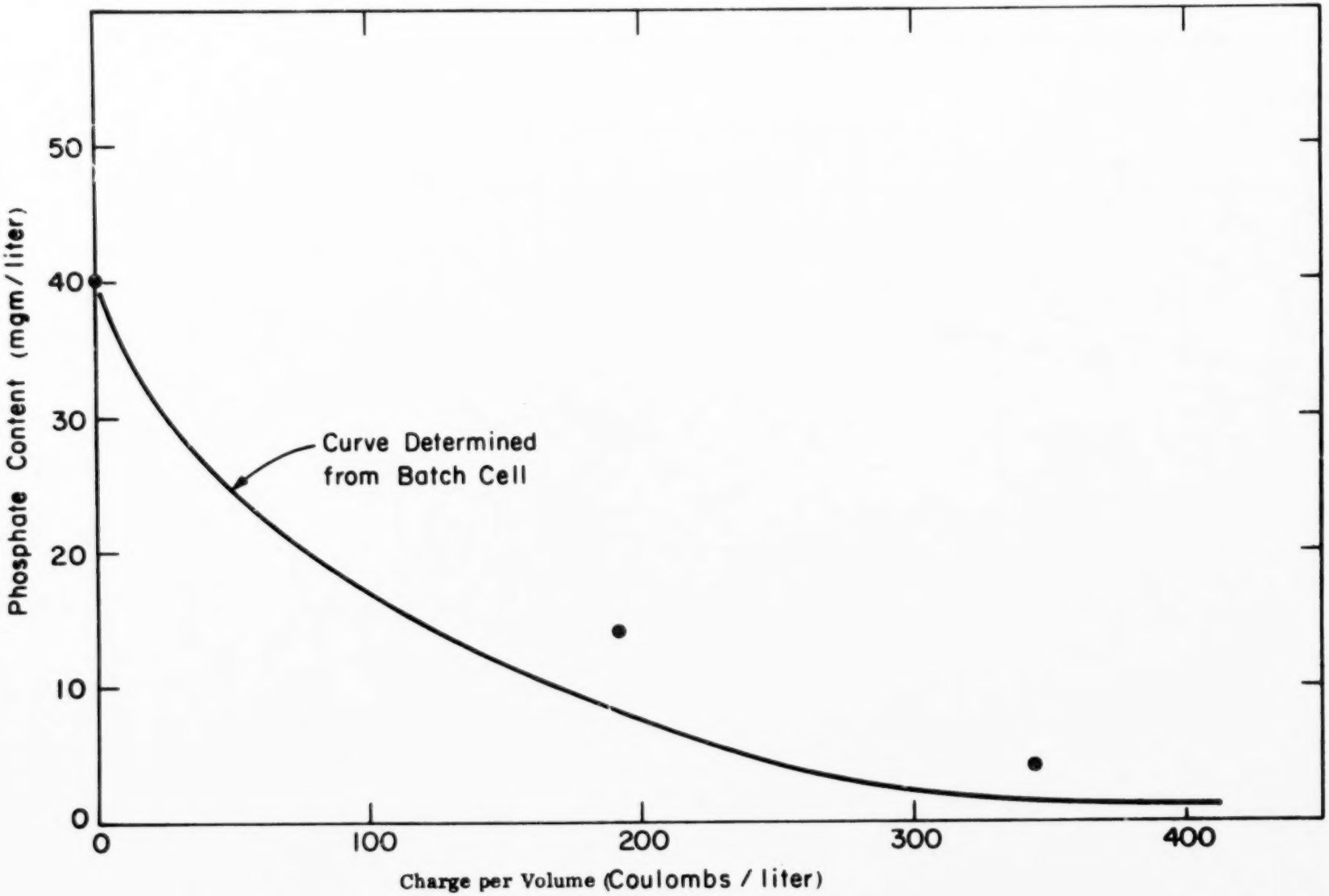

Fig. 17. Effect of Charge Flow Concentration on Phosphate Removal Using Aluminum Electrodes in Multiple Electrode Flow Cell

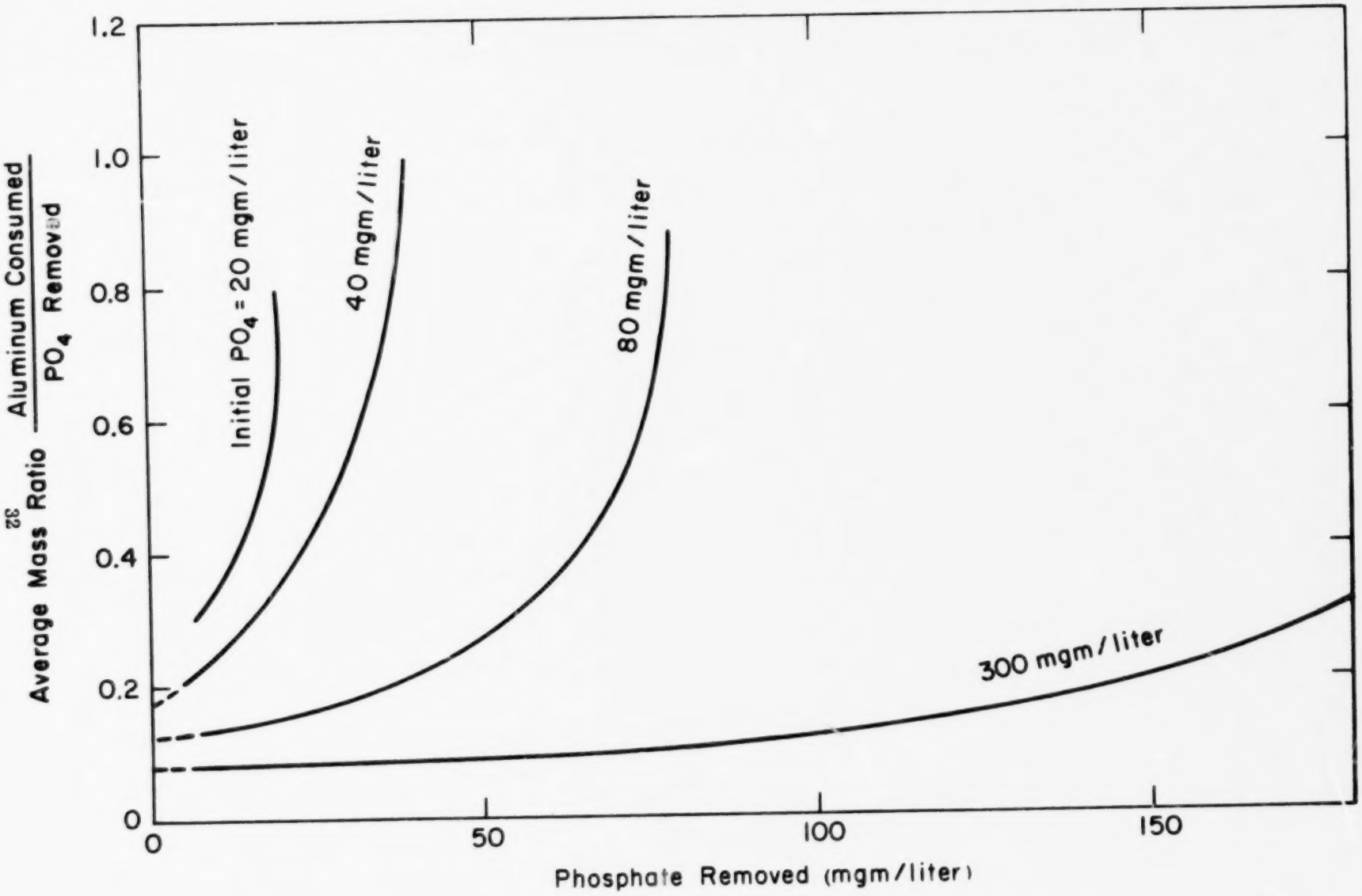

Fig. 18. Aluminum Electrode Consumption as a Function of Phosphate Removal

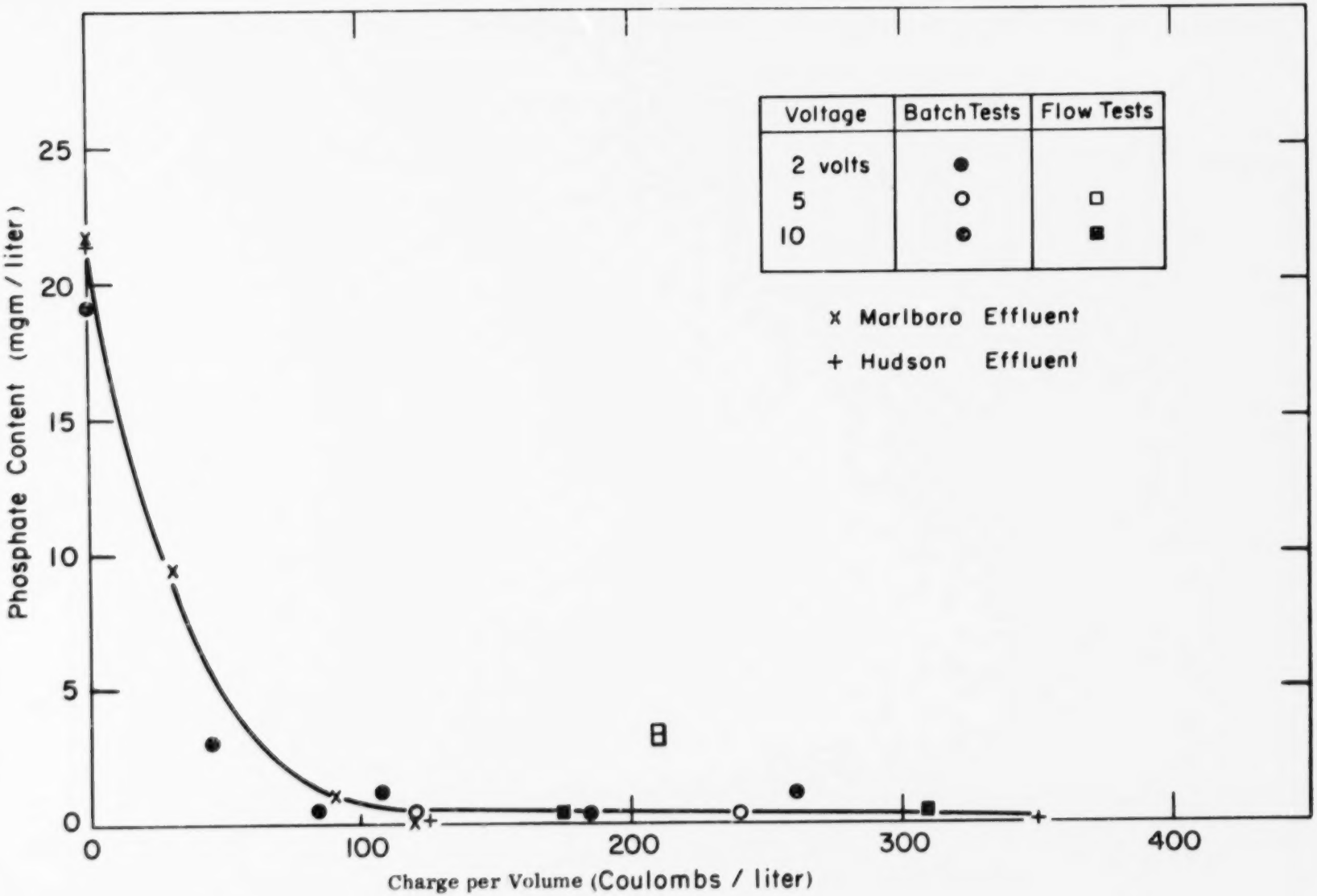

Fig. 19. Effect of Charge Flow Concentration on Phosphate Removal Using Iron Electrodes

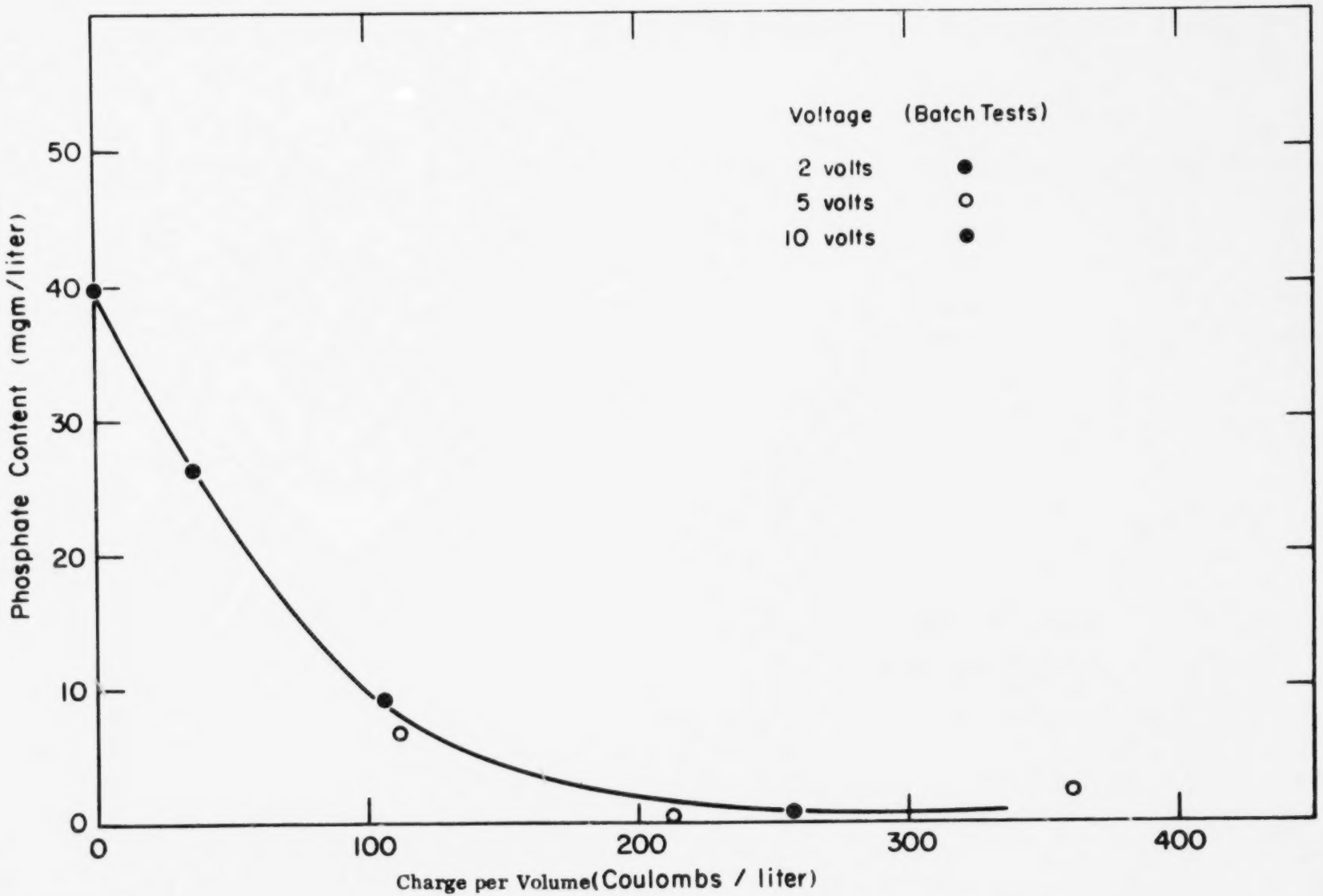

Fig. 20. Effect of Charge Flow Concentration on Phosphate Removal Using Iron Electrodes

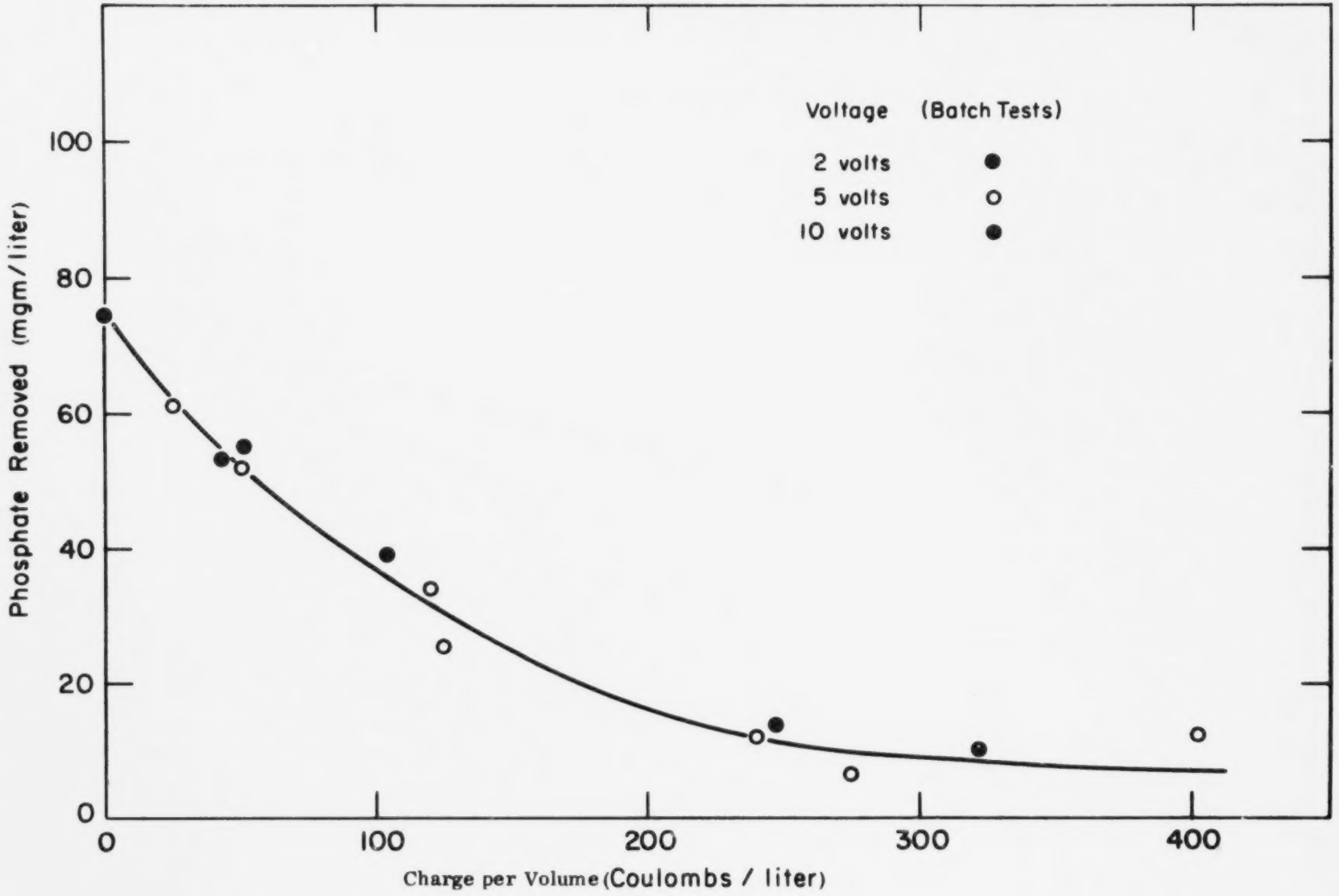

Fig. 21. Effect of Charge Flow Concentration on Phosphate Removal Using Iron Electrodes

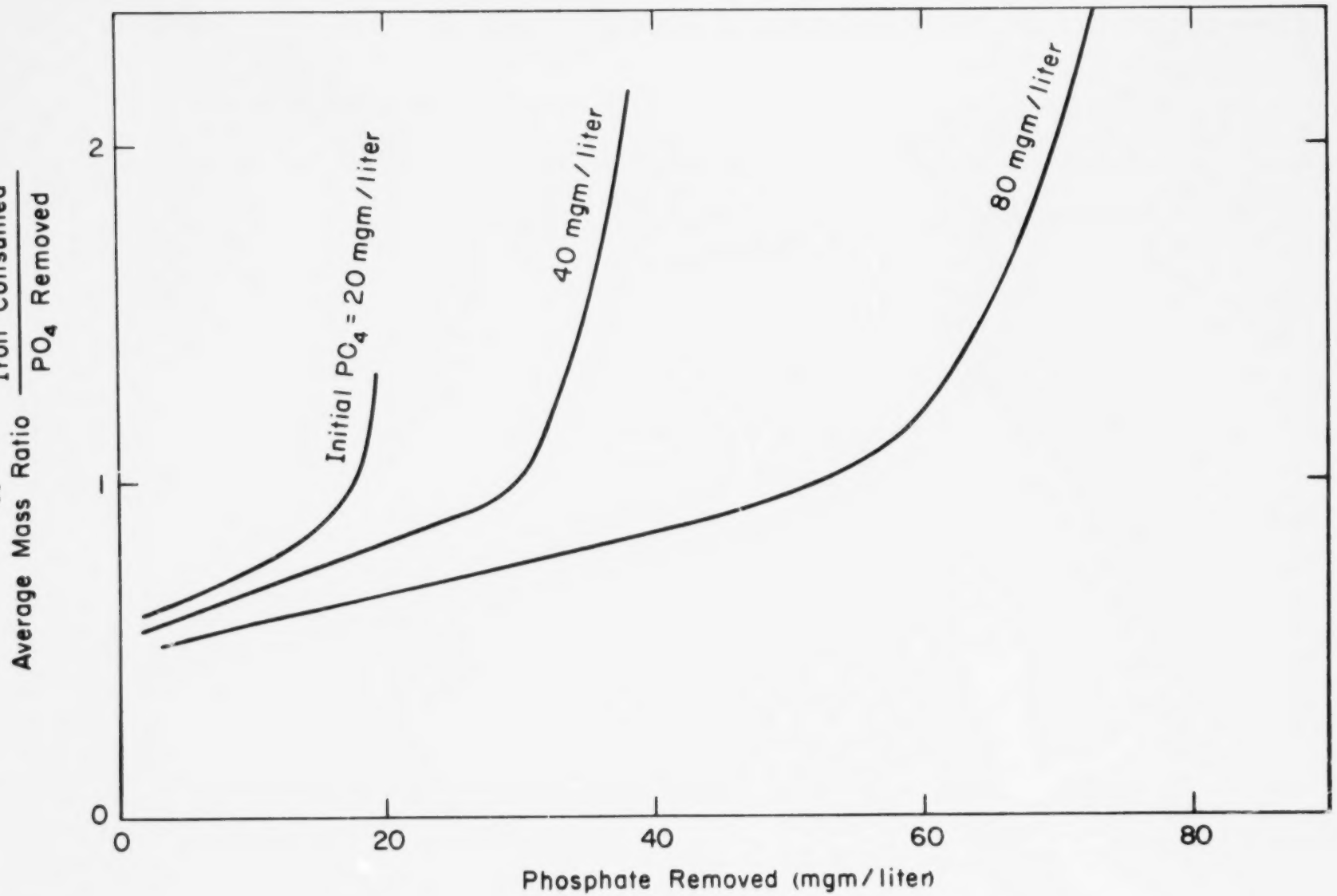

Fig. 22. Iron Electrode Consumption as a Function of Phosphate Removal

Table 6

RAW SEWAGE FLOTATION
(WARD STREET PUMPING STATION, BOSTON, MASS.)

	Voltage	Current (Amps)	Flow Rate (cc/min)	Coulombs / liter	Initial PO_4 mgm/liter		Final PO_4 mgm/liter	
					Total	Ortho	Total	Ortho
1	3	1.2	260	277	10.7	8.0	~2.7	0
2	3	1.2	360	200	10.6	8.0	~2.6	0
3*	4	1.4	360	233	11.0	8.0		0

*Measurements at the Cincinnati Water Research Laboratory on this set are as follows:

	Untreated Raw	Treated	
Suspended solids, mgm/liter	62.0	7.3	(mean of three measurements)
PO_4 - P, mgm/liter	4.5	0.36	"
TOC, mgm/liter	48.0	20.0	"

photograph of the cell operated with synthetic sewage shows that essentially all the precipitate is in fact carried to the top of the cell (Figure 23).

Table 6 indicates that in addition to phosphates, suspended solids are quite effectively reduced by the flotation. TOC content of the raw sewage is reduced from 48 to 20 mgm/liter.

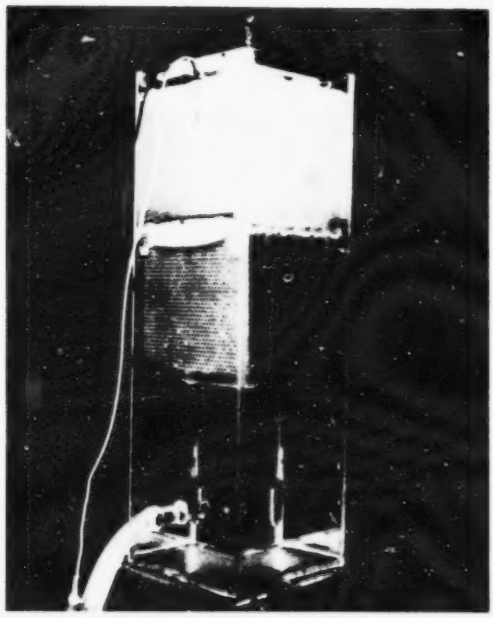

Figure 23

PHOTOGRAPH OF MULTIPLE ELECTRODE FLOW CELL
OPERATED WITH SYNTHETIC SEWAGE

SYSTEM OPTIMIZATION AND COST

Based on the data shown and the relationships developed in the previous section, one may proceed with a simple system optimization.

Such an optimization allows us to determine the dependency of the overall system design on variations in input parameter specification, and allows an approximate cost estimation to be made.

Assume that:

F = effluent treatment rate, gpd

A = dissolving electrode area, ft^2

s = electrode spacing, cm

V = voltage per cell, volts

σ = electrical conductivity of effluent, mho/cm

I = total current, amp (In the case of multiple electrode pairs, I would be divided by the number of these pairs.)

and that

η = current effectiveness = 0.2.

Therefore, the system volume = $A\,s/30.5\ ft^3$ (1)

and

$$I = V \eta \sigma \cdot \frac{A \times 930}{s} \text{ amps} \tag{2}$$

The charge flow concentration (Q) in coulombs/liter is then equal to:

$$Q = \frac{I \times 24 \times 3600}{F \times 3.785} \text{ coulombs/liter} \tag{3}$$

$$= 22{,}800\, I/F$$

Energy dissipation (E) = QV watt-sec/liter

$$E = 22{,}800 \frac{IV}{F} \text{ watt-sec/liter} \tag{4}$$

Now, the total system costs may be simplified and assumed to be broken down into three contributing parts:

1. energy cost
2. electrode consumption cost
3. equipment cost

Let

C_E = electricity cost, cent/watt-sec

C_M = electrode material cost, cent/lb

C_T = treatment tank cost, dollars/gal. capacity

W_e = equivalent weight of electrode material (9 for aluminum, 27 for iron)

z = electrode thickness, in.

ρ_M = electrode material density lb/ft^3

$$\text{Energy cost} = Q \cdot V \cdot C_E \quad \text{cents/liter (neglecting pumping power)} \tag{5}$$

$$\text{Electrode consumption cost} = \frac{Q \cdot (W_e) \times C_M}{96500 \times 454} \quad \text{cents/liter} \tag{6}$$

Equipment cost (neglecting transformers and rectifiers) is composed of tank cost and electrode interest costs.

$$\text{Tank cost} = \frac{100 \times C_T \times (As/4.08) \times 0.0614}{3.785F \times 360} \quad \text{cents/liter}$$

$$= \frac{C_T (As)}{906F} \quad \text{cents/liter} \tag{7}$$

and

$$\text{Electrode cost} = \frac{\rho_M A z C_M \times 0.0281}{12 \times 3.785F \times 360}$$

$$= \frac{\rho_M C_M A z}{581,900F} \quad \text{cents/liter}$$

where it is assumed that the equipment is amortized over 30 years at 4.5% interest and that operation is for 360 days per year.

Electrodes are assumed to have a resale value proportional to their residual weight.

The parameters I, F, and Q are interrelated by Eq. (3), while Eq. (2) relates V, A, I, and s. Using these two equations, the total cost may be expressed in terms of V and s after A is eliminated. This is:

$$\text{total cost} = Q \cdot V \cdot C_E + \frac{Q \cdot (W_e) \cdot C_M}{96500 \times 454} + \frac{Q\, C_T\, s^2}{906 \times 22800 \times 930\, V\, \eta\, \sigma} + \frac{Q\, \rho_M\, C_M\, s\, z}{581,900 \times 22,800 \times 930\, V\, \eta\, \sigma} \quad \text{cents/liter} \tag{8}$$

The cost is minimized at the lowest value of s and at an optimum value of the voltage V_o given by:

$$V_o^2 = \frac{s}{22800 \times 930\, \eta\, \sigma\, C_E} \left(\frac{s\, C_T}{906} + \frac{\rho_M\, z\, C_M}{581,900} \right)$$

It is interesting to note that the optimum voltage is not dependent on the plant capacity or the extent of the treatment required (phosphate removal).

For a typical effluent with a conductivity of 2×10^{-3} mhos/cm and a value of $\eta = 0.2$ with aluminum electrodes, the above equation reduces to

$$V_o^2 = \frac{10^{-7}\, s}{C_E} (1.3\, s\, C_T + 0.341\, z\, C_M)$$

Based on a tank cost of $\$100/yd^3$ (9), electrical energy costs of 1 cent/KWH and aluminum costs of 27 cents/lb (10)

$$C_T = \$0.49/\text{gallon}$$

$$C_E = 2.8 \times 10^{-7} \text{ cent/watt-sec}$$

$$C_M = 27 \text{ cent/lb.}$$

Therefore,

$$V_o^2 = 0.228\, s^2 + 3.29\, s\, z$$

Without considering pumping power it is impossible to do more than pick a reasonable value of s for these calculations. When a spacing of 3 cm and an electrode thickness of 0.5 in. are used, the optimum voltage is equal to 2.64 volts.

Furthermore, basing the calculation on a charge flow concentration of 300 coulombs/liter, the energy consumption is

$$E = 300 \times 2.64$$

$$= 792 \text{ watt-sec/liter}$$

$$= 0.835 \text{ KWH/1000 gallons.}$$

For a 1 million gallon per day plant

- Energy consumption 835 KWH/day
- Electrode area required $40,000\ ft^2$
- System volume $5,000\ ft^3$

  37,400 gallons
- Electrode consumption rate (with aluminum electrodes)

$$= \frac{300 \times 9}{96500 \times 454} \times 3.785 \times 10^6 \text{ lbs/day}$$

$$= 233 \text{ lbs/day}$$

This is equivalent to 1.4 ft^3/day. Assuming that the aluminum dissolves uniformly from the surface of the electrode, dissolution is equivalent to a reduction in thickness of 4.2×10^{-4} in./day. In one year, the aluminum consumption would be equivalent to a reduction in thickness of 0.3 in. The 0.5 in. figure assumed earlier would therefore require replacement about once per year.

Of the two operating costs, electrode consumption costs are significantly higher than electrical energy costs ($63/day for electrode vs. $8.4 day for power) when using aluminum.

When using iron, the equivalent electrode consumption is 700 lbs/day. At a cost of 2 cents/lb, this is equal to $14/day. Tables 7 and 8 summarize these results.

Table 7

ELECTROLYTIC PHOSPHATE REMOVAL SYSTEM DESIGN

Basis: 1 million gallon per day system

	Iron Electrodes	Al Electrodes
Voltage	1.76 volt	2.64 volt
Power Supply	23.2 KW	34.8 KW
Energy Required	0.557 KWH/1000 gallon	0.835 KWH/1000 gallon
Electrode Area	60,000 ft^2	40,000 ft^2
Tank Capacity	56,000 gallons	37,400 gallons
Electrode Consumption (mass)	700 lbs/day	233 lb/day
Electrode Consumption (volume)	1.5 ft^3/day	1.4 ft^3/day
Electrode Consumption (thickness)	0.30 mils/day	0.42 mils/day

NOTE: Electrode spacing assumed = 3 cm
Effluent conductivity = 2×10^{-3} mhos/cm
Phosphate removal > 95%

Table 8

ELECTROLYTIC PHOSPHATE REMOVAL CONCEPTUAL SYSTEM COSTS

	Iron Electrodes	Al Electrodes
Energy (at 1 cent/KWH)	0.56 cent/1000 gallon	0.84 cent/1000 gallon
Electrode	1.40 cent/1000 gallon (at $40/ton)	6.30 cent/1000 gallon (at 27 cent/lb)
Equipment Depreciation plus interest	0.56 cents/1000 gallon	0.84 cent/1000 gallon
TOTAL	2.52 cent/1000 gallon	7.98 cent/1000 gallon

These costs do not include labor, transformer, and rectifier costs, or filtration costs.

REFERENCES

1. Journal AWWA, Nov, 1965, p. 1431.
2. Journal WPCF, June, 1965, p. 800.
3. U. S. Patent 2,997,430 (Aug 22, 1961).
4. U. S. Patent 3,035,992 (May 22, 1962).
5. "Advances in Water Pollution Research," Proceedings of the International Conference held in London, Sept, 1962 (Vol. 2).
6. Chemical Engineering, Jan 4, 1965, p. 9.
7. AWTR-13, Public Health Service Publication No. 999-WP-13, March, 1965, "Electrochemical Treatment of Waste Water."
8. "Standard Methods for the Examination of Water and Waste Water," Prepared and published by APHA, AWWA, and WPCF, Twelfth Edition (APHA, N. Y.).
9. Perry's Chemical Engineers' Handbook, Fourth Ed., 1963, p. 26-21.
10. "Oil, Paint and Drug Reporter."

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